Gas equilibrium and exergy analysis of the FICFB and Viking gasification concepts

Martin Gassner*, François Maréchal
Laboratory for Industrial Energy Systems
Ecole Polytechnique Fédérale de Lausanne
CH – 1015 Lausanne, Switzerland

Abstract

Two biomass gasification concepts, i.e. indirectly heated, steam blown fluidised bed gasification and directly heated, air blown fixed bed gasification are modelled with adjusted equilibrium equations, their energy requirements are integrated and an exergy analysis of the different process sections is performed. The difference between the observed gas composition and the thermodynamic equilibrium is assessed, their overall performance is compared and major loss sources of the systems are identified.
Keywords: gasification, modelling, energy integration, exergy

1 Introduction

Suitable biomass gasification technology is usually identified considering design constraints like capacity and requirements like gas composition, calorific value and contaminants that are related to the specific application [1]. Process integration aspects are regarded to a lesser extent, although heat requirements of gasifiers are generally important and influence the systems' overall performance markedly.

Besides the traditional gasification application of power production in combined cycles [2] or gas engines [3, 4], the interest in fuel reforming processes based on biomass gasification currently increases [5, 6, 7, 8]. Since the primary product of such processes is the energy stored in the material outlet streams, a high chemical gas conversion is essential. Furthermore, the energy requirements of the reforming steps after gasification are different and proven technology established for power generation does not necessarily fit the modified demand. The purpose of this paper is to compare two potential gasification systems for fuel production with regard to their performance as gas generators. With this attempt, a further analysis considering the process integration with the fuel reforming steps as well as a process optimisation is prepared.

In particular, this work compares an indirectly heated, steam blown fluidised bed gasifier (FICFB case [3]) – being currently regarded as promising option for synthetic natural gas production [1] – with a directly heated, air blown fixed bed gasifier (Viking case [4]) with regard

*martin.gassner@epfl.ch, tel +41 21 693 53 16

to their performance as gas generators. In a first step, energy–flow models based on adjusted equilibrium equations are developed to investigate the level of thermodynamic conversion of the produced gas. The heat requirements of the gasification process are then calculated and allow to quantify the exergy losses of the different process sections. Finally, the energy integration of the process is performed to determine the overall thermodynamic perfomance of the system if its excess heat is used to produce power in a steam cycle.

2 Process modelling

2.1 Gasification reactor model

Due to the inherent need for simple model equations in flowsheet calculations, gasification systems are often modelled assuming thermodynamic equilibrium of the gas phase at reactor outlet. However, comparisons with experimental data show that this is generally not a valid assumption and kinetic effects must be taken into account. A convenient way to do so is to adjust the equilibrium equations in order to fit experimental data by introducing artificial temperature differences ΔT of the form [6, 7]:

$$\hat{K}_\mathrm{p} = K_\mathrm{p}(T_\mathrm{g} + \Delta T) \tag{1}$$

where K_p is the theoretical equilibrium constant and $\hat{K}_\mathrm{p}$ the equilibrium constant corresponding to the experimentally observed composition at the gasification temperature T_g. In this way, it is possible to develop a simple model that is valid with regard to both gas compositions and energy balances in the range of normal operation conditions.

Considering that the outlet streams of the gasfiers consist of H_2, CO, CO_2, CH_4, H_2O, N_2 and C(s), three equations in addition to the atomic balances are required to determine the gas composition. In our model, two solid–gas, i.e. the hydrogenating gasification (eq. 2) and the boudouard (eq. 3) equilibria, and the water–gas shift (eq. 4) reaction have been used:

$$C(s) + 2H_2 \rightleftharpoons CH_4 \qquad \Delta h_r^0 = -75 kJ/mol \tag{2}$$

$$C(s) + CO_2 \rightleftharpoons 2CO \qquad \Delta h_r^0 = +173 kJ/mol \tag{3}$$

$$CO + H_2O \rightleftharpoons CO_2 + H_2 \qquad \Delta h_r^0 = -41 kJ/mol \tag{4}$$

If experimental data of the gasifier shows that higher hydrocarbons are produced, its amount is calculated assuming that they are appropriately represented by ethene and that its partial pressure is proportional to one of methane:

$$p_\mathrm{C2H4} = k_\mathrm{p} p_\mathrm{CH4} \tag{5}$$

According to these equations, the four model constants are ΔT_hg (eq. 2), ΔT_bd (eq. 3), ΔT_wg (eq. 4) and k_p (eq. 5).

2.2 Energy–flow models

2.2.1 Flowsheet layout

Based on the flowsheets[1] shown in figure 1, energy–flow models for FICFB and Viking gasification are developed to compute the chemical and physical transformations and their corresponding heat requirements.

[1]The purpose of these schematics is to illustrate the model structure and does not correspond to the physical process layout, i.e. the pyrolysis, oxydation/combustion and gasification reactors correspond to one physical unit.

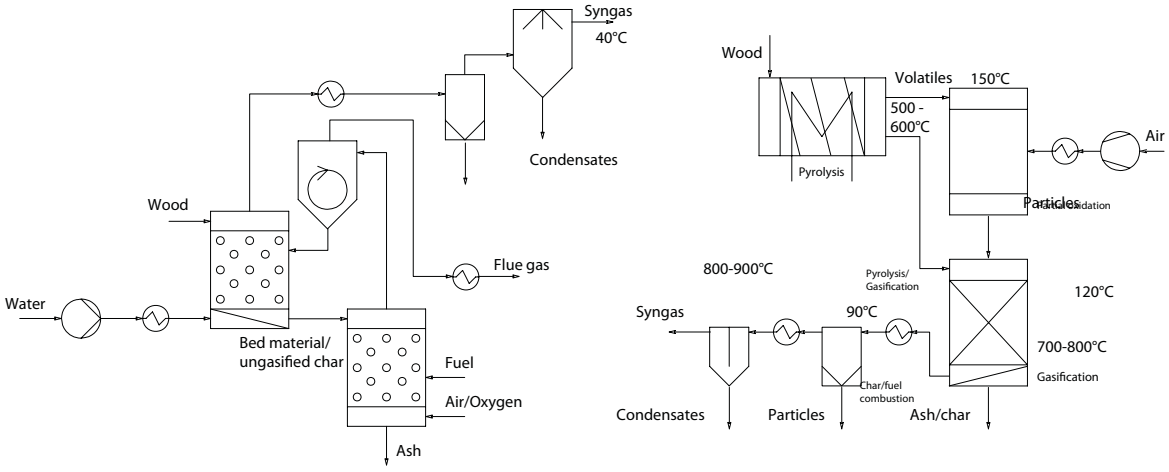

Figure 1: Conceptual flowsheets for FICFB (left) and Viking gasification.

In FICFB gasification, gasifying steam is injected into the fluidised bed where drying, pyrolysis and gasification of the raw material take place. The reactor is heated indirectly through circulating bed material by combustion of the ungasified char and additional fuel from a process stream. The obtained synthesis gas is cooled to 150°C, filtered and washed in order to remove dust particles, tar and other contaminants. In Viking gasification, wood is conveyed through a pyrolysis unit and heated to 500-600°C. After thermal cracking of the tars through partial oxidation of the gas phase, the remaining solid species are gasified in a fixed bed. The synthesis gas leaving the gasifier at 700-800°C is cooled to 90°C and filtered. Finally, the condensates are removed at ambient conditions in a gas-liquid separator.

2.2.2 Data reconciliation

With the general assumptions detailed in table 1, the model constants of the gasification reactor model are determined using data of wood characteristics, process conditions and gas compositions from [4, 9, 10]. For FICFB gasification, the reference temperature for gasification is of 850°C, where the artificial temperature differences in Viking gasification refer to pyrolysis and gasification temperatures of 600°C and 750°C respectively. The composition of the gas phase after partial oxidation is calculated assuming that the apparent equilibrium temperature after pyrolysis is unchanged. The values of the identified constants and their accuracy with measured data is given in tables 2 and 3.

2.2.3 Results and discussion

According to table 3, the model is able to reproduce the measured composition accurately. Except the CO_2 fraction of the FICFB reactor, all calculated compositions are within a 5% error range (5.5% for CO_2 of FICFB). However, no data for the gas composition after pyrolysis and partial oxidation are available for the Viking gasifier. Especially the calculated gas composition from pyrolysis is therefore in doubt. Experimental data of wood from a bench–scale system [11] report a dry gas composition of about 10%vol CH_4, 40%vol CO, 44%vol CO_2, 3%vol H_2 and 2%vol C_2. Furthermore, about 2/3 of the gaseous mass flow are represented by condensable hydrocarbons (tars), which are not considered by the model at all.

Parameter	Unit	FICFB	Viking
$\Delta p_{\text{gasifier}}$	mbar	150	50
Δp_{filter}	mbar	100	100
$\Delta p_{\text{exchangers}}$	mbar	-	-
$\eta_{\text{s,turbomachinery}}$	%	80	80
gasifier heat loss	%	10[1]	2[2]
wood humidity (inlet)[3]	%wt	10	30
volatiles[3]	%	-	84

[1] based on transferred heat
[2] based on LHV at inlet
[3] according to data used for reconciliation [9, 10]

Table 1: Model assumptions.

Process	FICFB	Viking	
Reactor	gasification	pyrolysis	gasification
ΔT_{hg}	-260°C	-289°C	-11°C
ΔT_{bd}	-201°C	-[1]	-123°C
ΔT_{wg}	-112°C	+12°C	-126°C
k_p	4.9	-	-

[1] It is assumed that the solid pyrolysis product contains not only char, but also hydrogen, oxygen and nitrogen with the same atomic distribution as wood. The solid outlet product is therefore controlled by the fraction of volatiles and reaction (3) is not used.

Table 2: Reconciliated model constants.

Process	FICFB		Viking			
Reactor	gasification		pyrolysis	partial oxidation	gasification	
State	wet	dry	wet	wet	wet	dry
CH_4	8.8 / 9.0	- / 9.3	- / 35.7	- / 3.2	- / 1.2	1.2 / 1.2
CO	29.4 / 28.0	- / 28.9	- / 3.0	- / 10.5	- / 18.3	19.6 / 19.0
CO_2	16.2 / 15.3	- / 15.9	- / 33.2	- / 16.6	- / 14.2	15.4 / 14.7
H_2	37.3 / 39.5	- / 41.0	- / 4.9	- / 21.6	- / 30.4	30.5 / 31.4
H_2O	3.6 / 3.5	- / -	- / 23.0	- / 13.6	- / 3.2	- / -
N_2	2.9 / 2.9	- / 3.0	- / 0.2	- / 34.5	- / 32.7	33.3 / 33.7
C_2H_4	1.8 / 1.8	- / 1.9	- / -	- / -	- / -	- / -

Table 3: Gasifier outlet compositions in %vol (Data [4, 10] / Calculation).

Examinating the reconciliated model constants, it is obvious that the synthesis gas from both gasifiers is not in thermodynamic equilibrium. Almost all apparent equilibrium temperatures are below the actual reactor temperatures. Compared to equilibrium, the gasifiers produce gases that contain too much hydrocarbons and carbon dioxide and too less hydrogen and carbon monoxide. The difference from equilibrium is thereby not as substantial in case of Viking gasification, which may be due to the different reactor types. A previous study [7] observed similar temperature deviations for air blown and steam blown gasifiers of fluidised bed type and reasoned similar catalytic activity of the bed. Unlike in fluidised bed reactors, the temperature in fixed beds is not equally distributed and part of the gas might be formed at higher temperatures. In case of Viking gasification, the pyrolysis gas is further heated to 1100°C by partial oxidation in order to thermally crack the tars. Since equilibrium further shifts towards H_2 and CO and reaction kinetics tend to accelerate with temperature, this might also lead to approach thermodynamic equilibrium at the gasifier outlet to a greater extent.

2.3 Energy integration

2.3.1 Problem definition

All process temperatures and heat requirements being calculated in the energy–flow models, the energy integration of the process is performed using the heat cascade concept. No heat exchanger network is defined prior to the identification of the minimum energy requirements. The calculated heat recovery system does thus not match the actual layout, but allows to determine the best solution with given utility and cogeneration possibilities.

In this analysis, the heat demand above the pinch is satisfied through combustion of gaseous process streams and, in the FICFB case, ungasified char. In directly heated gasification, residual char is diluted and lost in the ash. Since the process pinch point is at high temperature, power production from excess heat with a Rankine cycle is further possible. In this work, typical small scale operating conditions of 60 bar for steam production and 500°C for steam superheating are used.

2.3.2 Process performance

The performance of the process is assessed by calculating the energy balances, the cold gas efficiency $\varepsilon_{\mathrm{cg}}$ and the overall energy efficiency $\varepsilon_{\mathrm{tot}}$, defined as:

$$\varepsilon_{cg} = \frac{\Delta h^0_{i,gas} \dot{m}_{gas}}{\Delta h^0_{i,wood} \dot{m}_{wood}} \tag{6}$$

$$\varepsilon_{tot} = \frac{\Delta h^0_{i,gas} \dot{m}_{gas} + w^-}{\Delta h^0_{i,wood} \dot{m}_{wood} + w^+} \tag{7}$$

where Δh^0_i and $\dot{m}$ designate the lower heating value per unit mass and the mass flow respectively. w^- refers to overall produced and w^+ to overall consumed power, whereas only one of these terms unequal zero.

For a proper comparison of the gasification systems, the same feedstock properties (table 4) and an arbitrary total energy input of 1 $\mathrm{MW_{th,wood}}$ is chosen. The resulting energy balances and composite curves including the steam cycle for power recuperation are depicted in table 5 and figure 2. The plots reveal considerable differences of the performance of the investigated gasification concepts. Indirectly heated, steam blown gasification is pinched at the gasification

Proximate analysis		Ultimate analysis	
LHV	19.2 MJ/kg$_{dry}$	C	50.93 %wt
E_n[1]	21.6 MJ/kg$_{dry}$	H	6.11 %wt
Moisture	30.0 %wt	O	42.16 %wt
Ash content	0.5 %wt	N	0.80 %wt

[1] Chemical exergy is calculated according to [12].

Table 4: Feedstock properties.

Type		FICFB	Viking
Consumption	Wood	1000 kW	1000 kW
	Electricity	2 kW	2 kW
Production	Gas	748 kW	910 kW
	Electricity	68 kW	19 kW
Efficiency	ε_{cg}	74.8%	91.0%
	ε_{tot}	81.4%	92.8%

Table 5: Energy balances.

temperature which demands an important amount of fuel. Only 23% of the hot utility requirements of 265 kW at 875°C are provided by the combustion of ungasified char and 27% of the produced hot gas must be withdrawn from the process. If the clean and cold gas is used for this purpose, the withdrawn energy amounts to 24% of the totally produced gas and causes the cold gas and the total efficiency to decrease by 4.9% and 3.4% respectively. For the same reason, the amount of thermal energy available from the combustion gases below the pinch is elevated and 6.8% of the total wood input is recuperated by the steam turbine as mechanical work. In case of Viking gasification, the pinch point is not generated by the process stream but by the optimal heat recuperation network and no gas needs to be withdrawn and used as hot utility. In consequence, the cold gas efficiency is high and only few electrical energy is produced by the Rankine cycle.

3 Exergy analysis

3.1 Method

In order to properly identify the loss sources of the gasification systems, exergy balances considering the exergy value of material, thermal and mechanical streams (designated $\dot{E}_e$, $\dot{E}_q$ and $\dot{E}_w$ respectively) are defined for all process sections. The exergy losses $\dot{L}$ and efficiency η are then calculated according to:

$$\dot{L} = \sum \dot{E}_e^+ + \sum \dot{E}_q^+ + \sum \dot{E}_w^+ - \left(\sum \dot{E}_e^- + \sum \dot{E}_q^- + \sum \dot{E}_w^-\right) \quad (8)$$

$$\eta = 1 - \frac{\dot{L}}{\sum \dot{E}_e^+ + \sum \dot{E}_q^+ + \sum \dot{E}_w^+} \quad (9)$$

where superscripts '+' and '-' refer to streams entering and leaving the section, respectively. The exergy value of material streams is determined by adding its exergy value at atmospheric

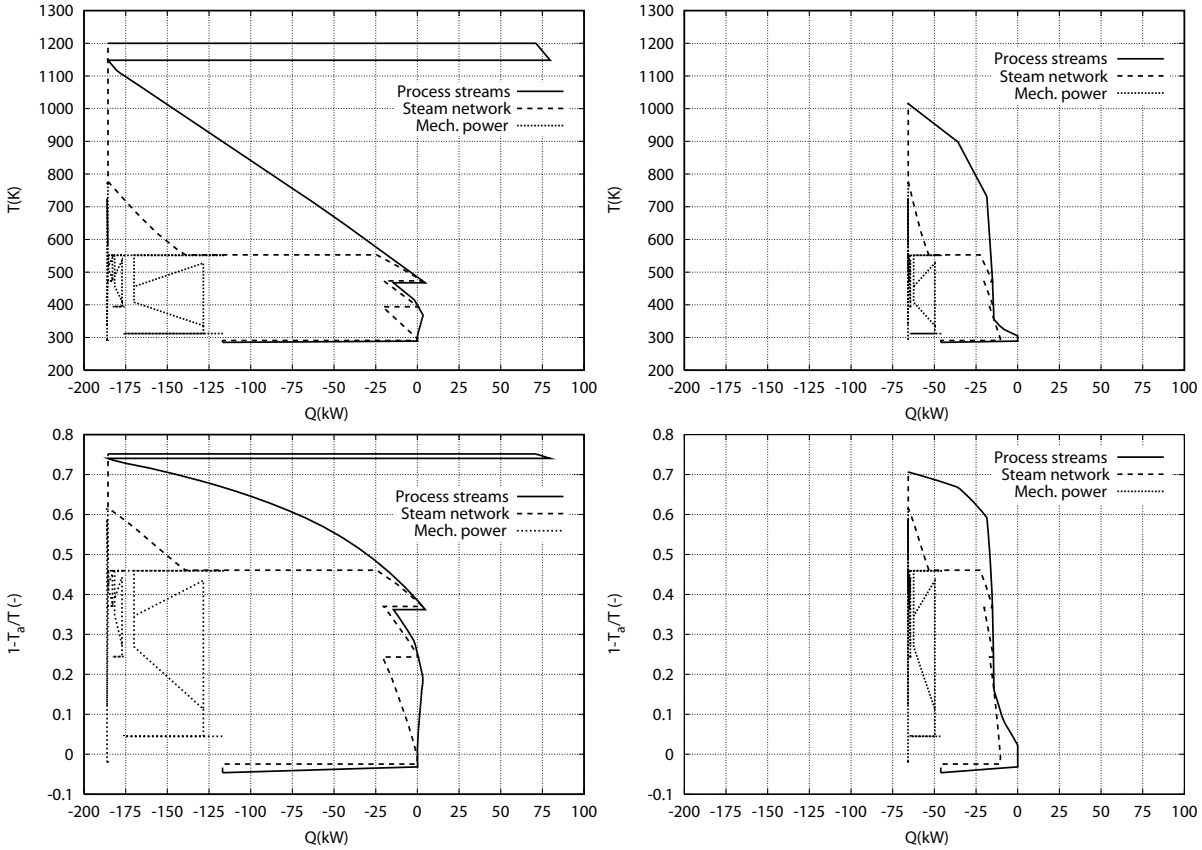

Figure 2: Energy (top) and exergy composite curves for FICFB (left) and Viking gasification.

temperature T_a and the exergy necessary for heating the stream:

$$\begin{aligned}\dot{E}_e &= \left[\Delta k^0 + \int_{T_a}^{T} c_p \left(1 - \frac{T_a}{T}\right) dT\right] \dot{m} \\ &= \left[\Delta k^0 + c_p \left(T - T_a(1 + \ln(T/T_a))\right)\right] \dot{m} \end{aligned} \tag{10}$$

In this equation, Δk^0 is the exergy value per unit mass and c_p the specific latent heat at constant pressure. The assumption of a constant c_p is also used to determine the exergy value of the thermal streams:

$$\begin{aligned}\dot{E}_\mathrm{q} &= \dot{m} \int_{T_\mathrm{a}}^{T} c_\mathrm{p} \left(1 - \frac{T_\mathrm{a}}{T}\right) dT \\ &= \dot{m} \cdot c_\mathrm{p} \left(T - T_\mathrm{a}(1 + \ln(T/T_\mathrm{a})\right) \end{aligned} \tag{11}$$

Finally, the exergy value $\dot{E}_w$ of mechanical streams equals their energy value w.

3.2 Results and discussion

As already indicated in the energy integration analysis, the exergy balances (table 6) of the process sections show that an efficient energy integration is crucial for a good thermodynamic overall performance. Although the exergy losses of the directly heated fixed bed gasifier exceed the ones of the FICFB gasifier due to mixing losses and partial oxidation, its total losses are considerably lower. In case of indirectly heated gasification, valuable thermal exergy is lost at high temperature. The pinch at 875°C and the fact that the recovery equipment only allows

Section	FICFB		Viking	
	$\dot{L}$ [kW]	η [%]	$\dot{L}$ [kW]	η [%]
Gasification[1]	175	87.4	245	80.2
Gas cooling[2]	17	75.4	11	89.7
Heat transfer	154	67.1	12	87.4
Steam turbine	16	84.1	5	82.0
Total	362	68.7	273	76.8

[1] including pyrolysis

[2] The exergy efficiency of the cooling section is assessed on the basis of the theoretical work potential of the rejected heat - the separation of condensates is not considered as a transformation and the chemical exergy of the inlet material stream is not included in the denominator of eq. 9.

Table 6: Exergy losses and efficiencies.

moderate operating conditions result in total heat transfer losses of 13.5% of the total exergy input. Since no additional combustion gases are produced, the amount of heat exchanged in directly heated gasification is considerably lower, and exergetic heat losses are restricted to 1% of the total process input. In this way, Viking gasification reaches an overall exergy efficiency of 76.8%, while FICFB gasification is limited to 68.7%.

4 Conclusion

Using adjusted equilibrium equations in flowsheet calcuations and performing the process integration with the heat cascade concept, indirectly heated, fluidised bed gasification with steam as gasifying agent and directly heated, fixed bed gasification with air as gasifying agent are compared with respect to the thermodynamic conversion and energetic performance. The modelling approach allows to accurately reproduce experimentally observed gas compositions and to quantify the difference to its thermodynamic equilibrium. It has been observed that the conversion in the directly heated fixed bed gasifier is closer to equilibrium and less hydrocarbons are formed. The reason for this is suspected in an unequal temperature distribution in the reactor and intermediate heating of the pyrolysis product for tar cracking.

In order to analyse the overall performance and identify the major loss sources, the energy integration has been completed with an exergy analysis of the different process sections. Especially due to a pinch of the FICFB process streams at high temperature, its hot utility requirements and exergy losses in heat transfer are elevated. The advantage of a nitrogen free product gas with high calorific value of this gasification technology is thus penalised by lower cold gas, overall energy and exergy efficiencies of 74.8%, 81.4% and 68.7% compared to 91.0%, 92.8% and 76.8% for Viking gasification, respectively.

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